GASIFICATION OF ORGANIC WASTE

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INTRODUCTION

The utilization of solid organic waste has received considerable
attention as a result of public concern over a deteriorating environ-
ment and an apparent depletion of our natural resources. Up to 70%
or 280 million tons of the 400 million tons of municipal and indus-
trial waste created annually can be classified as organic (1, 2).
This figure has been predicted to at least double by 1980 (3). How-
ever, these figures are small in comparison to the 3/4 billion tons of
moisture free organic animal waste produced annually (4). The problems
and expense associated with the disposal of these wastes have led inves-
tigators to seek new and unique solutions.

Among the oldest and currently most used disposal methods are land·
fill and incineration. Inherent problems associated with these two
methods have been described (5, 6). Methods have been devised and work
is continuing on the utilization of the heat created by incineration.
Other more innovative utilization processes have been proposed, some of
which are currently under investigation. The Bureau of Mines has been
investigating two different conversion processes--the pyrolysis method
and the hydrogenation process (7,8). The former method produces gas,
oil and solids while the latter, which appears to have a greater com-

mercial promise, produces essentially oil. Although these two processes are thought to have general applicability, they have largely employed bovine manure as a feed material. Two other processes, which are somewhat limited in scope, are Firestone's destructive distillation process (9) for converting shredded tires into various useful chemicals and Cities Service--Goodyear process (10) for the production of carbon black from old tires. In this communication the results of a bench-scale feasibility study on the conversion of solid organic waste into a high Btu fuel gas are presented. These results are discussed in reference to the process variables and the experimental equipment employed in the investigation.

EXPERIMENTAL

Reactor Design.--The gasification of the organic waste was carried out in a one inch, batch charge, bench-scale reactor. The schematic is shown in Figure 1 which has been previously described in detail (11). The reactor has been designed so that temperature, pressure and steam input can be controlled over a considerable range.

Feed Materials.--Several types of organic wastes were used as reactor feed materials. These include paper, tire rubber, Denver Municipal Sewage Sludge, animal waste and simulated garbage composed of 4.5% paper, 13.4% polyethylene, 39.5% organic waste, 24.7% tire rubber and 17.9% wood. The chemical analyses of these materials are given in Table I.

Catalysts.--Three different commercially available nickel catalysts were employed. Physical and chemical properties of these catalysts are
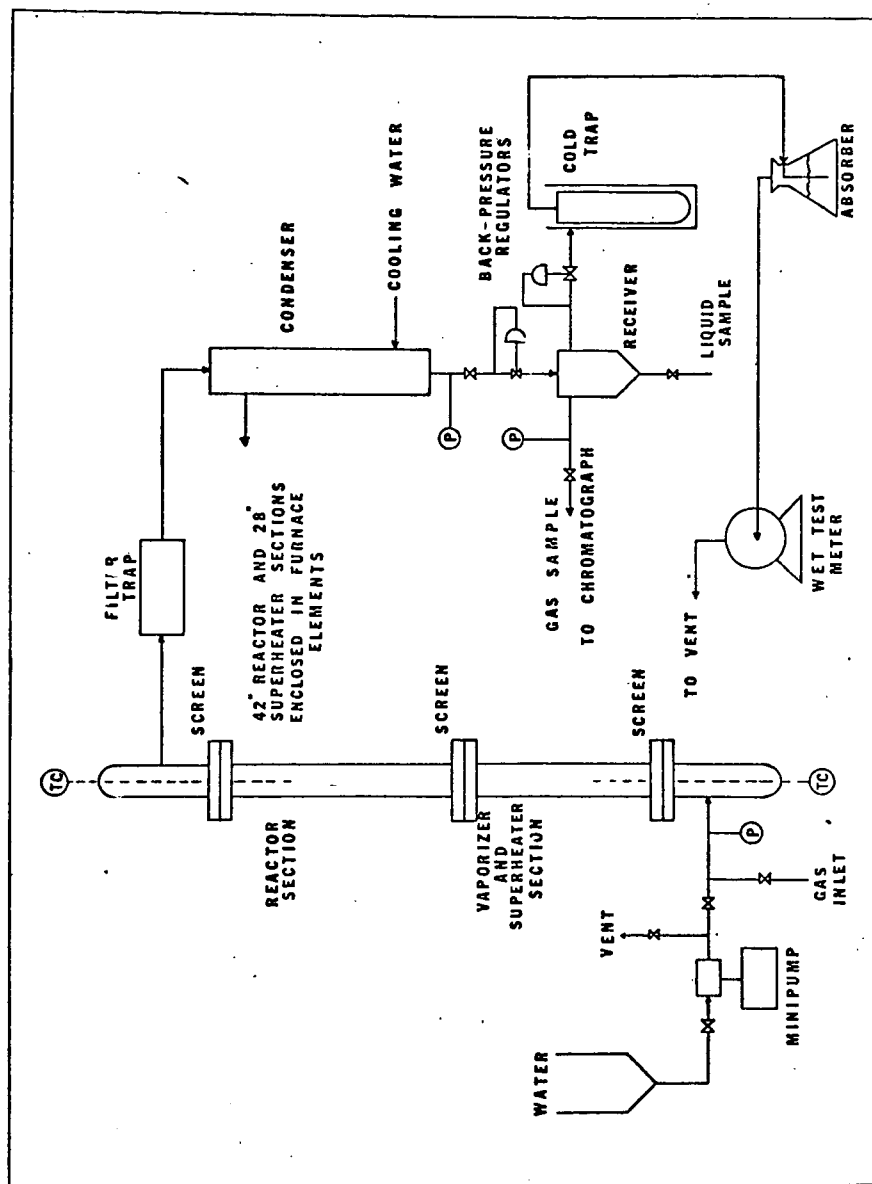
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Figure 1. Schematic Flow Diagram.
One Inch Reactor

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Table I. Analyses of Feed Materials

	Paper		Sewage Sludge		Scrap Tires	Charred Animal Waste	Raw Garbage Composite[a]
	Raw	Charred	Raw	Charred			
Moisture (wt %)	3.8	0.8	13.6	1.2	0.5		3.4
Hydrogen (wt %)	6.9	3.1	6.7	1.4	4.3	5.4	6.6
Carbon (wt %)	45.8	84.9	28.7	48.6	86.5	41.2	57.3
Nitrogen (wt %)	--	0.1	2.6	3.7	--	1.5	0.5
Oxygen	46.8	--	26.5	--	--	26.0	22.1
Sulfur (wt %)	0.1	0.1	0.6	--	1.2	0.4	0.4
Ash (wt %)	0.4	2.5	34.9	45.7	3.4	25.5	10.2

[a]Calculated from composite composition.

listed in Table II. In addition both potassium and sodium carbonates were employed as gasification catalysts.

Analyses.--Except as noted below, the majority of the chemical analyses were carried out by standard laboratory methods. Sulfur analyses were performed on a Leco sulfur analyzer and complimented by standard ASTM methods. Product gas volumes were recorded with a calibrated wet test meter. Product gas compositions were determined with a Beckman Model GC-5 duel column--duel thermal conductivity detector chromatograph equipped with an Auto Lab System IV digital integrator. Calibration values were obtained by use of a Linde primary standard gas mixture.

Methodology.--Experiments were carried out by charging the reactor section with a nickel catalyst and feed material. The feed material was mixed with an alkali carbonate catalyst. Both stratification (integrated) and separation (segregated) of the nickel catalyst and feed material were employed. The reactor was then adjusted to the desired operating conditions and the product gas was sampled for its composition every half hour. The compositions reported in Table III are the average compositions.

RESULTS AND DISCUSSION

It is apparent from the complex composition of the feed materials that the resulting gasification reactions are both varied and complex. Although these reactions may not be known from a mechanistic point of view, experimental observations allow one to formulate a functional description of the overall reactions which incidentally are consistent with those observed in the production of synthetic natural gas from coal.
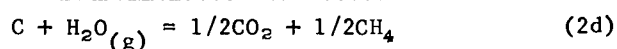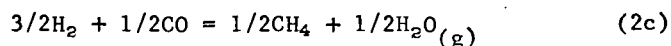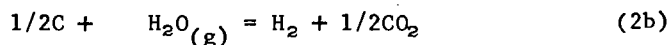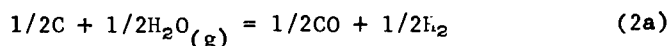
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Table II. Catalysts Properties [a]

	Harshaw Ni-0104	Harshaw Ni-3250	Girdler G-65
Total Ni (wt %)	58.8	50	25
Support Material	Kieselguhr	--	proprietary
Surface Area (m^2/g)	130	150	54
Pore Volume (cc/g)	0.5	0.34	0.08
ABD[b] (lbs/ft^3)	55	68	65
Pellet Shape and Size[c]	1/4 T	1/8 T	3/16 T
Avg. Crush Strength (lbs)	8	18	25

[a] Data is for a typical sample.

[b] ABD, apparent bulk density.

[c] Tablet, T. Sizes are expressed in inches.

$$C + H_2O_{(g)} = CO + H_2 \tag{1a}$$

$$\underline{H_2 + CO = 1/2CH_4 + 1/2CO_2} \tag{1b}$$

$$C + H_2O_{(g)} = 1/2CH_4 + 1/2CO_2 \tag{1c}$$

$$1/2C + 1/2H_2O_{(g)} = 1/2CO + 1/2H_2 \tag{2a}$$

$$1/2C + H_2O_{(g)} = H_2 + 1/2CO_2 \tag{2b}$$

$$\underline{3/2H_2 + 1/2CO = 1/2CH_4 + 1/2H_2O_{(g)}} \tag{2c}$$

$$C + H_2O_{(g)} = 1/2CO_2 + 1/2CH_4 \tag{2d}$$
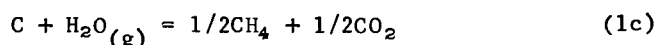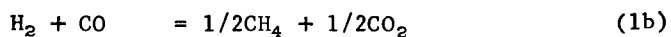
It should be noted that the end result in each of these reaction schemes is the same, irregardless of the nature of the intermediate steps. Employing these chemical equations in conjunction with the carbon content of the various organic wastes permits one to calculate a theoretical yield of gas product.

Table III contains a summary of some selected experimental gasification results for the various feed materials and catalysts employed. General observations obtained with the single stage experimental unit indicate that the initial reactions occurring during an experimental run are devolatilization and cracking, particularly in those instances where the volatile material content is high. This is followed by the carbon-steam reaction according to equation 1a which is initiated by the addition of steam. Alkali metal carbonates have been found to catalyze this reaction. Catalytic methanation of the intermediate reactants from the carbon-steam reaction with a nickel catalyst is then effected, apparently by reactions 1b and 2c.

Table III. Organic Waste Conversion Data

Run No.	302	295	489	416	298	297	517
Type Feed Material	Integrated Raw Paper	Integrated Charred Paper	Integrated Raw Sewage Sludge	Segregated Charred Sewage Sludge	Integrated Scrap tires	Integrated Charred Animal Waste	Integrated Raw Garbage Composite
Weight (1b)	0.132	0.172	0.275	0.275	0.207	0.220	0.240
Catalyst	Ni-0104	Ni-0104	Ni-3250	Ni-3250	Ni-0104	Ni-0104	G-65
Weight (1b)	0.165	0.191	0.220	0.275	0.220	0.220	0.275
K_2CO_3 (1b)	0.0330	0.0441	0.0331	0.0660	0.0551	0.0551	0.0440
Pressure (psia)	261	261	32	261	266	266	32
Ave. Temperature (°C)	710	740	670	390[b]	700	670	660
Gas Composition (mole %)							
CO	0.0	0.0	0.0	0.0	0.0	0.0	0.1
CO_2	48.0	42.7	44.8	48.9	49.3	48.2	42.0
H_2	8.8	10.4	3.2	6.7	1.5	5.0	4.2
C_1	43.2	47.5	52.0	44.5	49.3	46.9	53.7
unsats	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$>C_1$	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SCF CH_4/Ton	9,533	13,169	6,335	5,508	23,333	12,881	7,489
Theoretical CH_4 (SCF/Ton)[a]	12,823	25,410	6,917	6,822	27,320	13,012	17,150
CH_4 Yield (%)	74	52	92	81	85	99	44
Btu/SCF (CO_2-free)	884	883	973	923	981	937	963

[a]Based on equation 2d and carbon content of feed material.

[b]Apparent gasification temperature was ~720°C. Apparent catalyst temperature was ~390°C.

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The purpose of the addition of alkali carbonate to the feed material is to catalytically promote the carbon-steam reaction. This effects rates that would otherwise be prohibitively slow at these lower temperatures (650 - 750°C) in the absence of the catalyst. Potassium carbonate, sodium carbonate and sodium sesquicarbonate have been employed with no apparent chemical preference under experimental conditions. However, from an economic point of view the sodium sesqui-carbonate is the cheaper material. Although optimum alkali metal carbonate/carbon weight ratios have not been well defined, values between 12-25% have been found to give satisfactory results. These observations concerning the alkali carbonate catalysts are analogous to those found for the production of water gas from coal-steam reactions (12, 13).

The conversion of the intermediate reactants produced in the carbon-steam reaction are catalytically converted to methane, carbon dioxide and water by a methanation catalyst. Of the three catalysts employed, which have significantly different chemical compositions and physical properties, no apparent difference in their methanating ability under experimental conditions were observed. However, the catalyst's ability to produce significant quantities of methane from the initially produced gaseous reactants depends upon maintaining the nickel catalyst in a highly reduced state and protected from the air. This is achieved by reducing in situ and employing inert atmosphere techniques. In addition, the steam fed to the reactor must be judicially controlled in order to preserve the methanation catalyst's activity, yet provide enough steam to maintain a reasonable carbon-steam reaction rate. Excessive amounts

of steam will cause the gas composition to shift from low H_2, high CO_2 and CH_4 to high H_2, CO_2 and moderate amounts of CO. This shift is apparently due to reaction 2b as a result of the increase in the partial pressure of steam and possible catalyst oxidation. This CO-shift is accompanied by a drastic reduction in the volume heating value of the product gas.

The overall gasification process was found to be independent of pressure under experimental conditions. Thus, even though pressures of 261 psia were employed the results were essentially the same as obtained at 32 psia. This observation is consistent with theoretical predictions based upon equation 1c for an integrated reactor.

The temperature of the gasification process is a compromise between that necessary for the carbon-steam reaction and that for optimum methanation. This compromise is necessary when working with a one-stage process as opposed to a two-stage process. In the latter instance the temperature of each stage (carbon-steam and methanation reactions) can be adjusted for optimization. Depending upon concentrations of the gaseous reactants, the optimum temperature lies in the 590 - 675°C range for a single-stage process which is consistent with some thermodynamic calculations for the system (14).

Irrespective of the organic waste material employed, the gasification results were generally above 75% of the theoretical methane value. The quantity of gaseous product was proportional to the carbon content when other factors were held constant. Hence, as demonstrated by Table III, rubber tires which contain about 85 wt % carbon produced 23,000 SCF

CH_4/ton while Denver metropolitan sewage sludge with 26 wt % carbon yielded only 6300 SCF CH_4/ton. The sewage sludge was unusual from the standpoint that liquid hydrocarbons were also produced from its gasification. This liquid hydrocarbon production is believed to be attributed to the iron content (6 wt %) in the sludge which is introduced during municipal treatment. Iron has long been known to function as a Fisher-Tropsch catalyst.

Table IV contains a tabulation of volume contraction of the organic waste upon gasification. These values are based on apparent bulk densities for each of the feed materials. Also included within the table is a comparison between the theoretical potential energy value of the feed material that would be realized through combustion and that obtained through its conversion to a synthetic natural gas. The potential energy value of the synthetic natural gas is observed to be considerably less than that for combustion. This is largely a consequence of the chemical reactions involved in the two different processes. Furthermore, it must be realized that the efficiency of the conversion of the heat energy, created by incineration, into a usable and transportable energy form will decrease the difference between the two. Fernandes and Cohan (15) have computed the energy available from the incineration of mixed refuse containing 27% carbon and 4.5% hydrogen to be 3150 Btu/lb.

CONCLUSIONS

Exploratory experiments have shown that organic waste materials can be converted to gaseous fuel products composed essentially of CH_4, CO_2 and H_2 with a CO_2-free heating value in excess of 900 Btu/SCF. Furthermore, the overall conversion has been shown to be pressure

Table IV. Volume Contractions and Potential Energy Comparisons

Gasified Material	Run No.	% Volume Contraction	Calculated Heating Value of Reactant Material (Btu/lb reactant)	Experimental Heating Value (CO_2-free) of Gas Product (Btu/lb reactant)	% Potential Energy Recovered
Raw Paper	320	93.9	10020	5120	51
Charred Paper	295	62.2	13560	7090	52
Raw Sewage Sludge	489	61.9	7500	3270	44
Charred Sewage Sludge	416	43.1	7570	2920	39
Scrap Tires	298	83.3	14410	11780	82
Charred Animal Waste	297	63.4	8590	6680	78
Raw Garbage Composite	517	66.4	11480	3890	34

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independent over the range of 32-261 psia while the optimum temperature

(650 - 750°C) for the described process is strongly dependent upon the

composition of the primary gaseous reactants produced from the steam-

carbon reaction which is catalyzed by the presence of alkali carbonates.

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HYDROGASIFICATION OF CATTLE MANURE TO PIPELINE GAS

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INTRODUCTION

Growing shortages of pipeline gas and the increasing need to improve methods of waste disposal in order to minimize environmental problems prompted the Bureau of Mines to investigate the feasibility of converting solid wastes to pipeline quality gas by direct hydrogasification. A feasibility study has been published[1,2] that indicates that this means of disposing of municipal solid waste may, for larger municipalities, not only provide the lowest cost means of disposing of the solid waste but also provide a supplementary source of pipeline gas at a price lower than supplementary gas from other sources.

Of all the forms of organic solid waste, the most abundant is animal manure which amounts to at least 200 MM tons/yr of organic solids.[3] A major contributor to this manure is the modern beef cattle industry and it is in this industry that the pollution problems arising from manure disposal are most extreme. This is because the increasing size and concentration of modern feed-lots itensify manure disposal and pollution problems which range from water pollution arising from soluble nitrogen compounds in the manure to severe odor problems. But on the other hand, it is the increasing size and concentration of these feed-lots which make cattle manure a potentially attractive feed stock for pipeline gas plants because collection is simplified and larger scale plants, which allow the lowest unit cost production of pipeline gas, can be built. In addition, the areas with great concentration of feed lots such as West Texas are also areas where convenient gas transmission pipelines already exist.

In this report, experimental data are presented showing the quality and yield of pipeline gas that can be generated by directly reacting cow manure with hydrogen at gasification conditions.

EXPERIMENTAL

Procedure

Except for one experiment conducted with dried cow manure in a continuous free-fall dilute-phase reactor, the experiments with manure and solid wastes were conducted in a batch autoclave. A drawing of the assembled reactor is shown in figure 1. The autoclave body is fitted with a pyrex glass liner into which the autoclave charge is placed. A thermocouple well covered by a pyrex glass tube is inserted into the liner. The free volume of the assembled autoclave is 1.2 liters; the liner has a volume of 0.7 liters.

The autoclave, containing the charge, is assembled and weighed on a bullion balance. The autoclave is then installed in the electric furnace. The system is first purged with nitrogen to remove the air and then purged with hydrogen to remove the nitrogen. Oxygen in the hydrogen (usually 0.1 percent or less) is removed by passage through a vessel containing palladium on an alumina carrier. The oxygen is catalytically reduced to water which in turn is removed by passage through a vessel packed with anhydrous calcium sulfate (Drierite). The autoclave is then charged with hydrogen to the desired

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initial pressure. Selection of the initial pressure is governed by the pressure, at the specified reaction temperature, or by the hydrogen/solid feed ratio desired.

Upon reaching the specified pressure in the autoclave, the valve on the autoclave is closed and the rest of the system is purged with nitrogen to remove the hydrogen. The autoclave is disconnected from the purged system.

Rotation of the autoclave is started at the same time the furnace is energized. The heating rate is 8° C per minute. Reaction time is that period during which the temperature is levelled off at a specified value. For all the experiments reported here, the time at reaction temperature was one hour. Cooling is accelerated by removing the upper section of the furnace. Rotation is continued until the autoclave has cooled to 250° C or lower. The cooling rate is about 4° C per minute down to 250° C.

When the autoclave has cooled to room temperature, it is depressurized at approximately 3 scf/hr. The gas in the autoclave is first passed through a Nesbitt absorption bulb containing Drierite to remove water. The gas is then metered and several samples taken to determine the gas composition with a gas chromatograph. The absorption bulb containing Drierite is weighed to determine the yield of water and the char remaining in the autoclave is weighed and analyzed for carbon and hydrogen so that an overall material balance as well as carbon and hydrogen balances can be made.

Results With Cow Manure

Operating results obtained from the direct hydrogasification of dried cow manure are tabulated in table 1. Ultimate analysis of the manure used in these experiments is given in table 2.

The purpose of these experiments was to establish the feasibility of converting manure to pipeline gas and to determine the operating conditions at which this conversion could best be carried out. These results indicate that manure is an excellent feed stock for pipeline gas synthesis. For example, in runs MH-1 and MH-2 a product gas having a heating value in excess of 1,000 Btu/scf was produced after simply scrubbing out the CO_2. Thus, it is possible to produce an excellent pipeline gas from manure without methanation. That this is possible is due to the high concentrations of ethane in the product gases from the manure hydrogasification. The ethane is present in such high concentrations because the high reactivity of the manure allows hydrogasification to take place at low enough temperatures for the ethane to survive in the presence of hydrogen. At the higher temperatures necessary to hydrogasify coal, most of the ethane formed is quickly converted to methane by the reaction

$$C_2H_6 + H_2 \rightarrow 2CH_4.$$
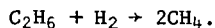
One can visualize alternate processing options available for the raw gases from manure hydrogasification. One option would be simply to shift the CO to CO_2 by the well-known water-gas shift reaction ($CO + H_2O \rightleftarrows CO_2 + H_2$) to get the CO level down to acceptable pipeline standards and then scrub out the CO_2. The resulting gas could then be placed directly in a pipeline without requiring any methanation. Another alternative would be to first separate the ethane, which could be sold separately and/or blended with the pipeline gas. In this case, the pipeline gas would be produced by methanating a mixture of CO and CO_2. The concentration of CO_2 would be adjusted by varying the amount of CO_2 removed during scrubbing. These options are now being investigated in a detailed process study which will be published when complete.

Effect of Process Variables

The primary independent variables studied in the batch autoclave were the hydrogen/manure feed ratio and the reactor temperature. Since the reactions are carried out in

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TABLE 1.- Summary of autoclave results for the direct
hydrogasification of cow manure

	MH-1	MH-2	MH-3	MH-4	MH-5
Reactor temp., ° C	550	550	650	550	475
Initial press., psig	150	150	150	150	150
Operat. press., psig	1800	1630	2200	750	1500
Solid charge, grams	80	80	80	20	80
Hydrogen charge, g-moles....	.562	.562	.562	.562	.562
Solid residue, grams	38.8	40.6	37.0	9.3	39.9
Carbon in charge, g-moles...	2.363	2.363	2.363	.591	2.363
Gas produced, g-moles.......	.963	.967	1.14	.65	.811
Carbon gasified, %	39.9	40.4	41.4	50.1	27.0
Water recovered, g-mole	.925	.872	.967	.233	1.083

Gas Analyses (Dry and N_2-Free*)

	MH-1	MH-2	MH-3	MH-4	MH-5
H_2	13.1	11.3	14.4	60.2	30.2
CO	.8	.7	1.7	4.1	.9
CO_2	34.1	33.5	29.3	10.6	38.7
CH_4	40.3	42.1	54.2	18.2	21.7
C_2H_6	11.7	12.4	.4	6.6	6.5
H_2S	-			.3	.2
C_3H_8	-				1.8
Total	100.0	100.0	100.0	100.0	100.0

*Reported N_2-free because the variable amounts (usually about1%
of N_2 were due to N_2 from purging sample lines. The nitrogen
content of the waste is converted to NH_3 which is dissolved in
water.

TABLE 2.- Ultimate analysis of dried cow manure

	Wt Pct
Carbon	35.4
Hydrogen	4.2
Nitrogen	0.7
Sulfur	0.2
Oxygen	23.5 (by diff.)
Ash	36.0
Total ...	100.0
Moisture ...	2.5

a batch reactor, setting the feed ratio and the reactor temperature determines, among other things, the pressure that the reactor reaches at the desired temperature. While only a few batch experiments were made, they were sufficient to establish fairly accurately the optimum temperature and roughly the range of hydrogen/manure feed ratios over which it would be reasonable to operate a plant converting manure to pipeline gas.

Temperature

Experiments were conducted at temperatures of 475,550 and 650° C. The most dramatic effect of temperature was on the fraction of carbon gasified and on the ethane yield. These trends are shown in figure 2 at a constant H_2/manure feed ratio of .007 g-moles H_2/gram manure. In addition, a single point is also shown at a H_2/manure ratio of .028 g-moles/gram in figure 1. As figure 2 indicates, in the batch reactor operating temperatures above 550° C are detrimental because the only effect of the increased temperature is to convert ethane to methane which consumes hydrogen and which lowers the overall thermal efficiency of the process. At this point, it should be noted that in a balanced plant in which the hydrogen is internally produced all the carbon in the manure is utilized. This requires that enough carbon should be left in the char from the hydrogasification step to produce the hydrogen. Preliminary calculations indicate that the conversion of about 40 percent of the carbon fed to the hydrogasifier will leave sufficient carbon to produce the hydrogen required for the hydrogasification step.

The water yield was not greatly affected by variations in temperature over the range studied and amounted to about 20 wt pct of the manure charged. Since the manure as charged had a moisture content of only 2.5 wt pct, most of the water was formed by release of bound water or by reaction of oxygen in the manure with hydrogen.

Water was the only liquid product formed in these tests. No tars or oils were formed in any of the experiments. Thus, the disposal of or recycle of tars should not be a problem in a plant converting manure to pipeline gas.

Hydrogen/Manure Feed Ratio

Four of the five tests conducted with manure were made at a hydrogen/manure feed ratio of .007 g-moles H_2/gram-manure with a single test at an increased ratio of .028 g-moles/gram. As the data in table 1 and the single point on figure 2 indicate, increasing the hydrogen/manure feed ratio resulted in a substantial increase in carbon conversion and the yield of ethane with a less significant increase in methane yield. However, at the higher hydrogen/methane feed ratio, both carbon monoxide and dioxide would have to be methanated to produce pipeline quality gas. It is worthwhile noting that in spite of the operating pressure being considerably less in MH-4 than in the other tests, because of the greater void space in the reactor due to the smaller solid change, the hydrogen partial pressure in this test was still substantially higher than in the other tests at 550° C. However, even with this increase in hydrogen partial pressure of about 2.5 times that in the lower hydrogen/manure ratio tests at 550° F, the overall carbon conversion or methane production were not increased anywhere near in proportion to the increase in hydrogen partial pressure.

Hydrogasification of Manure in a Continuous Reactor

Because of the promising results hydrogasifying dried cattle manure in a batch reactor, a hydrogasification test was made in a free-fall dilute-phase (FDP) reactor used for our coal hydrogasification studies. This FDP reactor and the method of conducting tests in it have been described.[4] To maximize the information from the FDP

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reactor run, the wall temperature was changed over the course of the experiment and the response of the gas composition to the different reactor wall temperatures measured. Data from these experiments are summarized in figure 3. As figure 3 indicates in the continuous FDP reactor, the methane yield increased greatly with increasing reactor wall temperature over the range of reactor wall temperatures studied. As in the batch tests illustrated in figure 2, the ethane yield in the FDP reactor went through a maximum because, as previously mentioned, with increasing reactor wall temperature, the ratio of the (rate of ethane disappearance by hydrogenation to methane) to the (ethane formation rate) increases. In the batch tests increased methane yields with increasing temperature above 550° C were due to the hydrogenation of ethane rather than the hydrogenation of the manure. However, in the continuous tests almost all of the methane was generated from the manure over the entire range of reactor wall temperatures.

It is difficult to make comparisons between the results in the FDP reactor and those in the batch reactor because the solid residence time in the FDP reactor is, at most, a couple of seconds compared to the hour residence time at temperature in the batch reactor tests. In addition, in the FDP reactor the solids are fed in cold and heat up very rapidly as they pass through the reactor and, because the time at any one temperature is very short, it is difficult to speak of a particle temperature. The physical appearance of the manure after hydrogasification was greatly different in batch and FDP experiments. Particles from the batch reactor had the same physical appearance as the feed manure except for being blackened while those from the FDP reactor were spheres with a very porous interior indicating that they had softened upon their passage through the reactor. In appearance, they seem almost identical to the char formed by the hydrogasification of raw coal.[4] In reference 4 it was pointed out that the high reactivity of the raw coal with hydrogen in the dilute phase is due to a great extent to the porosity of the particle which exposes the entire particle volume to hydrogen. The almost identical appearance of the hydrogasified manure particles indicates a similar mechanism applies when manure reacts with hydrogen in the dilute phase. Greater hydrogen accessibility greatly increases the reactivity of the manure as evidenced by the steadily increasing carbon conversion with increasing reactor wall temperature shown in figure 3 contrasted with the leveling off of carbon conversion with increasing temperature in the batch experiments shown in figure 2. The differences in the physical structure of the particles after reaction in the batch autoclave and the FDP reactor are due to two factors. The first is the difference in heatup rates that the particles in the two types of reactors experience. In the batch autoclave, the particles take approximately an hour to reach a typical reaction temperature while, in the FDP reactor, the heatup time is on the order of tenths of a second. The second factor is the difference in particle concentration between the two types of reactors. In the batch reactor the particles are in contact with each other or the reactor walls and the void fraction is typical of that for a fixed bed. However, in the dilute phase the particles are freely falling and widely separated with a void fraction approximately 95 percent. Thus, the combination of shock heating and unconstrained expansion result in a highly expanded porous particle that is readily permeated by hydrogen and hence extremely reactive.

Since in large-scale continuous reactor systems, the particle heatup rate will be the same order of magnitude as the heatup rates in the FDP reactor, one can expect the solids from such systems to be more reactive than the solids formed in the batch exploratory tests reported here. The use of this batch data for preliminary process design is therefore probably conservative.

SUMMARY AND CONCLUSIONS

Batch experiments show that cattle manure is readily converted to pipeline gas by hydrogasification at temperatures low enough to allow appreciable yields of ethane. It

is possible to produce a SNG with a heating value in excess of 1,000 Btu/scf by simply hydrogasifying the manure, shifting a rather low (on the order of 1 to 2 vol pct) concentration of CO to CO_2, and scrubbing out the CO_2 without any need for methanation. In addition, it was found that no tars or oils were produced from manure hydrogasification in spite of the relatively low temperatures. An experiment made with cattle manure in a continuous free-fall dilute-phase reactor indicated that the manure in such a reactor system is more reactive than in the batch reactor because of the much higher heatup rates and the low concentration of particles in the dilute-phase reactor.

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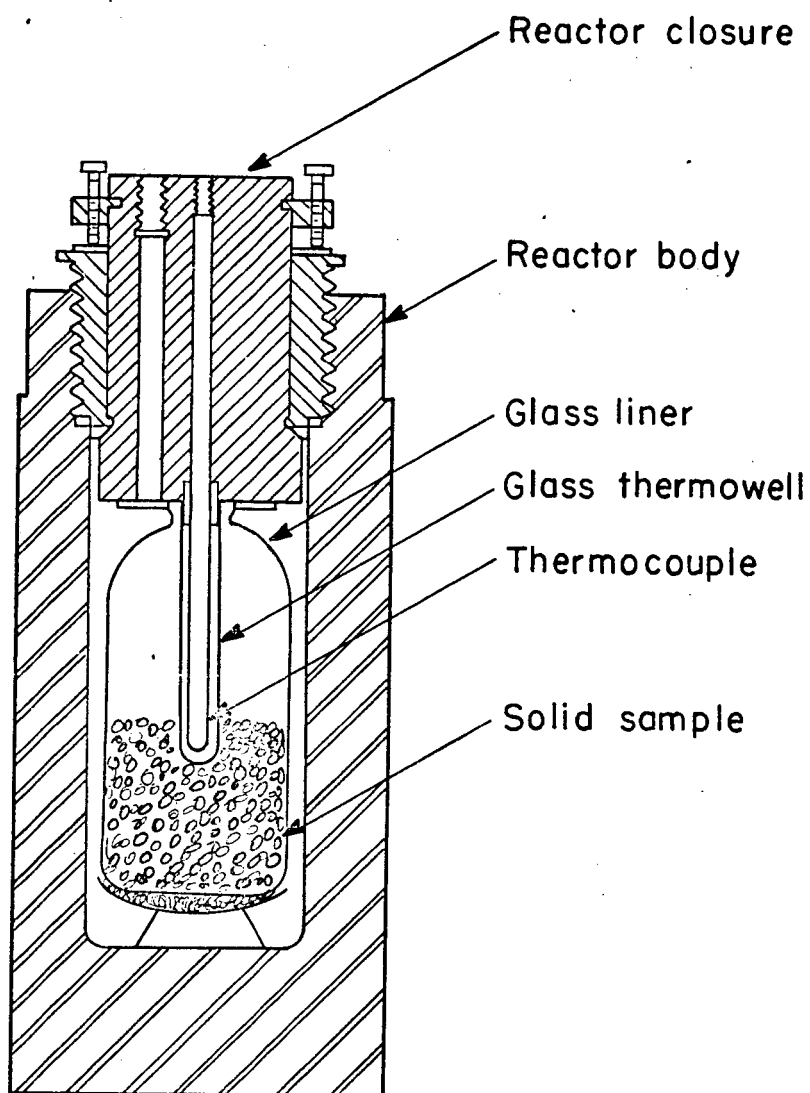
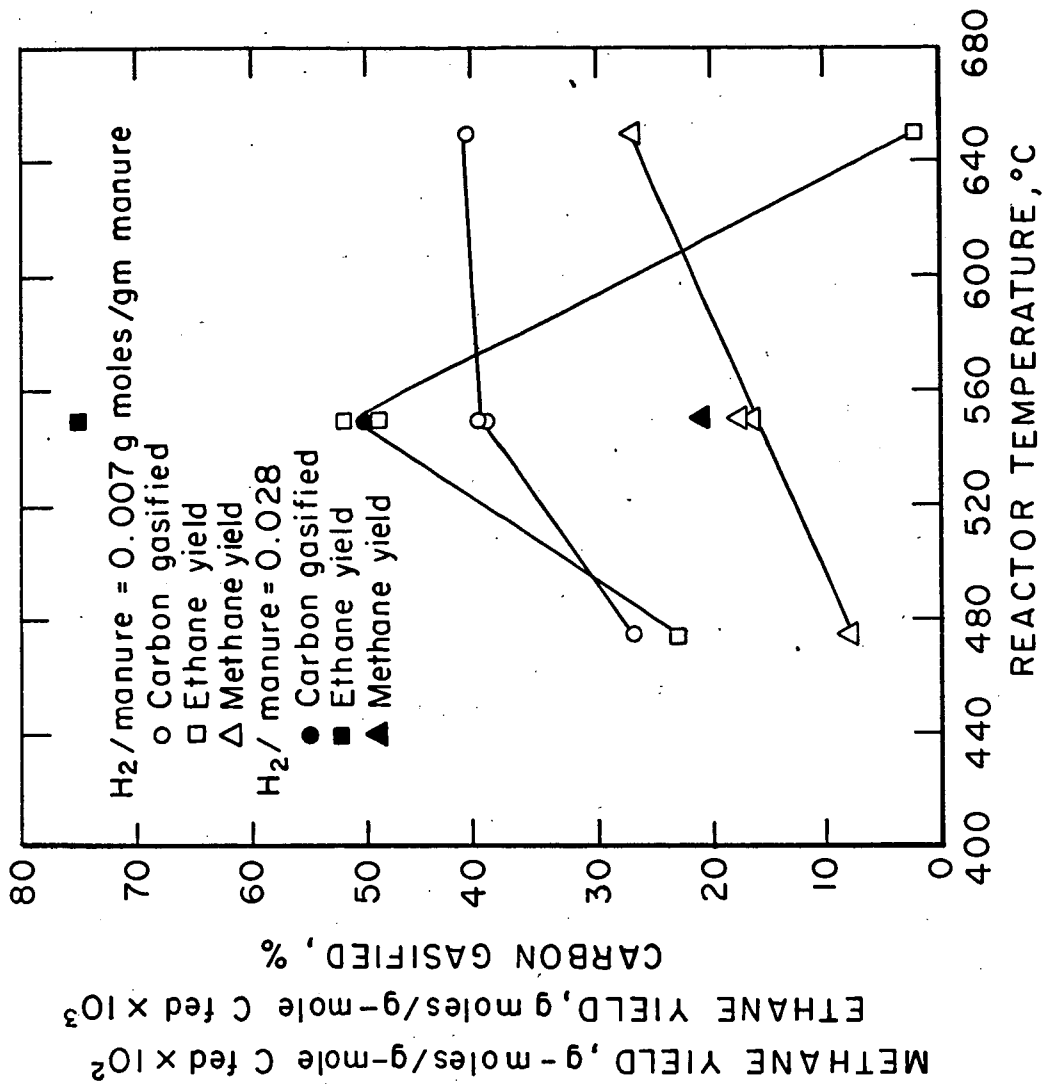Figure I — Schematic diagram of autoclave.

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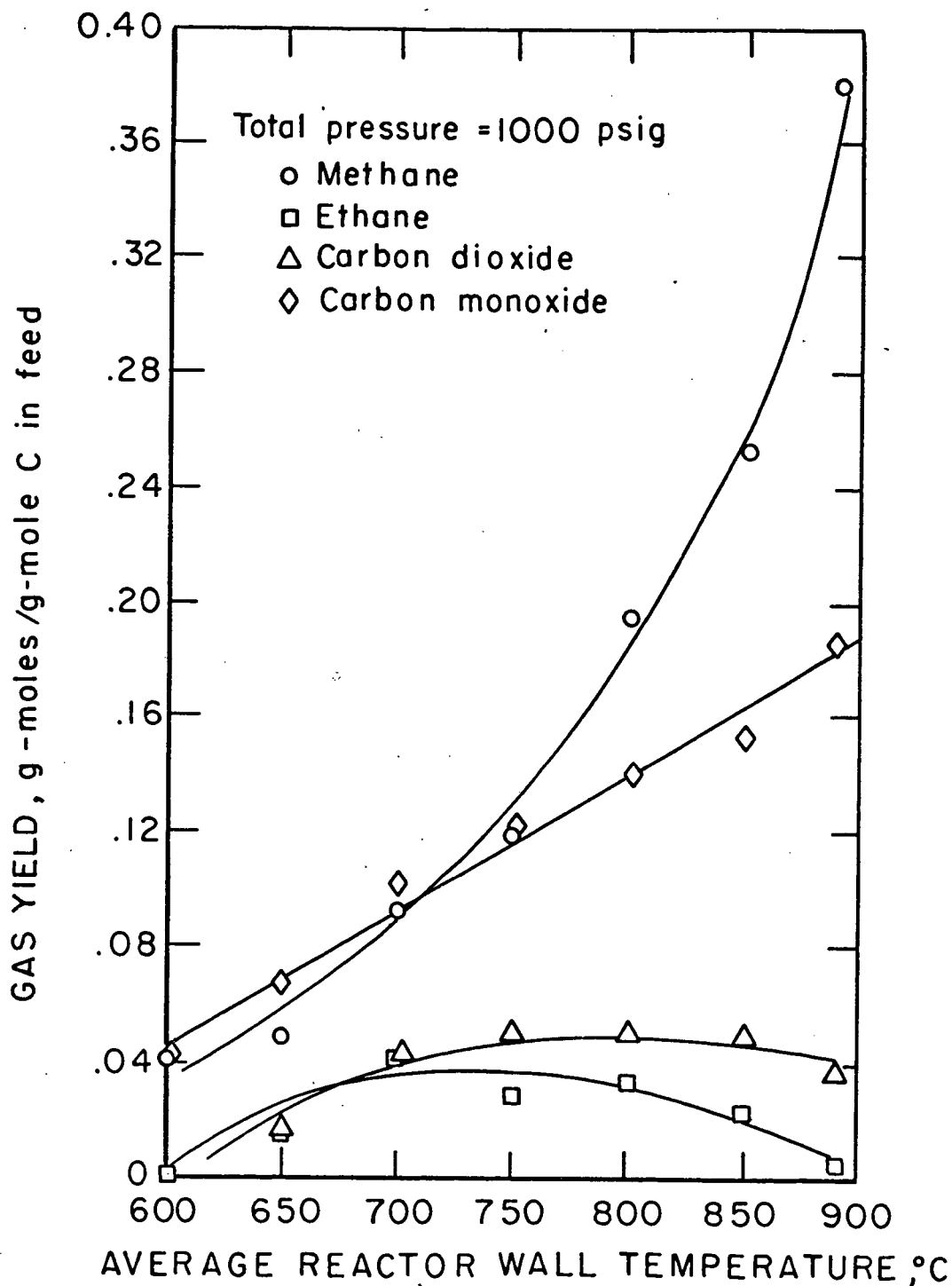Figure 2— Effect of reactor temperature on carbon conversion and hydrocarbon yield.

Figure 3— Effect of reactor wall temperature on hydrogasification gas yields from dried cattle manure.

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FLUIDIZED BED COMBUSTION OF MUNICIPAL SOLID WASTE IN THE CPU-400 PILOT PLANT.

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An overall concept for solid waste management known as the CPU-400 is currently under development at Combustion Power Company, under contract to the Environmental Protection Agency. The CPU-400 will utilize a fluidized bed combustor in a gas turbine cycle to convert the heating value of municipal solid waste into electricity. This paper describes the pilot plant and early results of experiments conducted on a low pressure configuration of the fluidized bed combustor and supporting equipment. The latter includes an inertial separator used to remove exhaust gas particles and a solid waste handling subsystem to prepare the municipal solid waste. Properties of the shredded, air-classified solid waste fuel are presented and combustor performance is discussed. The flow of inert material through the system during a recent 35 hour duration test is also presented along with results from a set of exhaust gas sampling instruments.

PILOT PLANT STUDY OF A VORTEX INCINERATOR FOR MUNICIPAL WASTE, C. H. Schwartz, A. A. Orning, R.B. Snedden, J.J. Demeter, and D. Bienstock, Bureau of Mines, 4800 Forbes Ave., Pittsburgh, Pa. 15213.

The incineration of a "standard" refuse and a selected municipal refuse from which most of the glass and metal had been separated was studied in a pilot scale vortex incinerator. Designed for a capacity of 150 to 300 pounds of refuse per hour, the incinerator has a refractory lining, a mechanical ram for continuous feeding, tangential-air firing, and provision for injecting additional air to the center of the vortex. Air, admitted tangentially at the top of the incinerator follows a helical path along the wall to the bottom where it contacts the bed of burning refuse. Gaseous combustion products, at high temperature and correspondingly low density, spiral up through the vortex to the furnace outlet at the top. Incomplete combustion experienced earlier was corrected by admitting excess air in the center of the hot ascending gases where temperature ranged from 2200°F immediately above the "fuel" bed to 1600°F at the top of the chamber. Auxiliary fuel (natural gas) was required to stabilize combustion when the moisture content of the municipal garbage exceeded 50%. Particulates in the effluent were 0.07 to 0.99 gr/cu ft for the "standard" refuse and 0.05 to 0.15 gr/cu ft for the municipal refuse. Combustibles in the stack ash ranged from 2.1 to 3.5%.